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ACTIVATED SLUDGE TREATMENT SYSTEMS WITH OXYGEN

John B. Stamberg, et al

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Abstract: The gas-tight biological reactor with 2.5 hour detention time or less insolubilized the biodegradable organics to less than 5 mg/l of soluble BOD. The organisms in the mixed liquor were maintained between 4000 and 8000 mg/l. Above an SRT of six days, the rate of activity in the oxygen volatile solids was greater than in a parallel step aeration system. Ninety-five (95) percent of the oxygen supplied was consistently ~~utilized~~ in the staged reactor, which employed co-current liquid and gas flow contacting. solids retention time

The liquid solids separation was accomplished by conventional clarification. The clarification efficiency was a function of mixed liquor concentration, particle shape, particle density and seasonal variation (i.e., temperature, metabolic changes, and load variation). During the warmer temperature periods, the peak overflow rates of 1940 gal/day/ft^2 were observed. In colder temperature periods, maximum steady state overflow rates of 975 gal/day/ft^2 were obtained. The underflow solids varied with overflow rate, clarifier volume and recycle rate. Underflow concentrations up to 2.5% were obtained.

Above an SRT of six days, the total production of solids was significantly less than the solids production in a similarly operated step aeration system. At an SRT of 13 days, 0.35 lb. of solids/lb. of BOD were produced.

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September 1973

ACTIVATED SLUDCE TREATMENT SYSTEMS WITH OXYGEN

by

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Alan B. Hais

Contract No. 14-12-818
Project 11010 EYM
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ABSTRACT

The oxygen activated sludge system must be viewed as an entirely unique approach, and compared on a total system basis with other alternative systems. The system is composed of three interrrelated subsystems; a biological reactor, a clarifier, and a solids handling system.

The unique gas-tight biological reactor normally with less than 2.5 hours of aeration time essentially completely insolubilized the biodegradable organics to less than 5 mg/l of soluble BOD in the Washington, D.C. wastewater. In D.C., the organisms in the mixed liquor were maintained between 4000 and 8000 mg/l and were visually similar to a conventional system except that the rate of activity in the oxygen volatile solids was greater than a conventional system above an SRT of 6 days. Greater than 95% of the oxygen supplied was consistently utilized in the staged reactor which employed co-current liquid and gas flow contacting.

The liquid solids separation was accomplished by conventional clarification. The clarification efficiency was a function of mixed liquor concentration, particle shape, particle density and seasonal variation (i.e., temperature, metabolic changes, and load variation). During the warmer temperature periods, the peak overflow rates of 1940 gal/day/ft^2 were observed. In colder temperature periods, steady state overflow rates 975 gal/day/ft^2 were the maximum obtained. The underflow solids varied with overflow rate, clarifier volume and recycle rate. Under flow concentrations up to 2.5% were obtained.

The total production of solids was significantly less than the similarly operated diffused air system above an SRT of 6 days. As little as 0.35 lb. of solids/lb. of BOD were produced at an SRT of 13 days.

This report was submitted in partial fulfillment of Project 11010 EYM and Contract No. 14-12-818 by the Department of Environmental Services, Government of the District of Columbia under the sponsorship of the Environmental Protection Agency. Work was completed as of September 1971.

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SECTION I

CONCLUSIONS

1. A gas-tight biological oxygen reactor with independent control of dissolved oxygen and mixing, coupled with an aerobic final clarifier, produces a good quality secondary effluent on District of Columbia primary effluent with 1.5 to 2.5 hours average detention time (based on raw flow) with MLSS concentrations between 4000 and 8000 mg/l.

2. Biodegradable organics in the District of Columbia primary effluent are essential completely insolubilized by the oxygen process (less than 5 mg/l of soluble BOD). Total carbonaceous BOD removal depends upon the amount of suspended solids in the effluent and, therefore, on the ability to clarify.

3. Oxygen micro-organisms are visually the same as those in a typical conventional system; however, the rate of activity of the oxygen volatile solids is greater above an SRT of 6 days.

4. Oxygen activated sludge is subject to filamentous (Sphaerotilis) growth as similarly observed in the air systems when operated below an SRT of 5 days on D.C. primary effluent.

5. Sludge in the oxygen system underflow settles to approximately 1.0-1.4% solids in a clarifier with 1.9 hours of hydraulic retention time and 2.0-2.4% with 2.8 hours.

6. Total production of excess biological solids is significantly lower in the District of Columbia oxygen system than in a parallel step aeration system at SRT's above 6 days with as little as 0.35 lb. of excess solids produced/lb. of BOD added at an SRT of 13 days.

7. When the oxygen clarifier is operated with a deep feed well or with the mixed liquor sufficiently concentrated to settle in a subsidence (zone) settling pattern, the blanket acted as a filter and produced high quality effluent. In the summer and fall, 1970, the clarifier operated at a peak rate of 1940 gpd/ft^2. In the 1970-71 winter, oxygen clarifier rates could not exceed a sustained 975 gpd/ft^2. A larger clarification area is required in the winter than in the Summer on District of Columbia wastewater for a given MLSS concentration for the same effluent quality.

8. With a shallow center feed well and with the mixed liquor concentrations low enough (under 4500 mg/l) to permit discrete particle settling, better settling rates are observed in the oxygen clarifier than with the method of clarifier operation described in No. 7 above. Only moderate decreases in effluent quality (increase in

SS from 15 to 25 mg/l) in consecutive tests with similar operating conditions are observed with this type of clarifier operation at the District of Columbia.

9. Substantial nitrification is achieved in the oxygen aeration system in the summer and fall at the District of Columbia.

10. Average effluent phosphorus residuals of 1.8 mg/l as P with an alum dosage of 1.4 Al^{+++}/P, by weight, were achieved in the oxygen system. Higher phosphorus removals are possible with higher alum dosages, but in areas with moderate wastewater alkalinity, pH control may be required to prevent the depletion of wastewater alkalinity reserves in the oxygen system.

11. Based upon the influent and exhaust gas flows, over 95% of the input oxygen is consistently utilized in the District of Columbia oxygen reactor.

12. The oxygen activated sludge system must be viewed as an entirely unique approach, and compared on a total system basis with other alternative systems. Reactor and clarifier sizing must be coordinated. As the reactor size is increased, a low MLSS concentration is required for a given biological state (F/M ratio). The lower the MLSS concentration, the smaller the required secondary clarification area. The solids handling requirement of an oxygen system depend on the biology established in the reactor/clarifier combination. Thus, if minimum excess biological sludge production is required, then more capacity is required in the reactor/clarifier combination. Further, the concentration of the clarifier underflow solids is dependent on clarifier volume and recycle rate.

SECTION II

RECOMMENDATIONS

1. The oxygen activated sludge system fed with raw wastewater should be evaluated over a wide range of operation conditions.

2. The oxygen activated sludge system, being fed either raw or primary wastewater, should be evaluated as a high rate system with high F/M ratios and lower SRT's than employed in this study.

3. Mineral addition (alum) for phosphorus removal should be further evaluated especially using lime to restore pH and alkalinity.

4. Other methods of influent feeding such as step aeration and contact stabilization should be evaluated with oxygen aeration.

5. The use of separate oxygen digestion of primary, secondary or a combination of waste sludges should be investigated as a potential system alternative.

6. A comprehensive thickening dewatering study is required on a variety of equipment to define the best probable alternatives to handle the waste oxygen solids.

7. Alternate liquid solids separation methods to gravity clarification should be considered.

8. A study on designing of full scale circular and rectangular clarifiers from pilot plant and batch test information is badly needed not only for oxygen, but for all biological sludges.

9. Separate nitrification utilizing oxygen should be evaluated.

SECTION III

INTRODUCTION

The use of pure oxygen within the activated sludge process dates back some 20 years. Pirnie (1) proposed a system of predissolving pure oxygen in high concentrations in the influent wastewater before entering a non-aerated mixed reactor. The process was termed "a bio precipitation system." Biological success was achieved by Okun (2) in bench scale tests and later by Budd and Lambeth (3) on a pilot scale, but oxygen utilization efficiencies of 20-25% were too costly. Okun and Lynn (4) and later Okun (5) showed an increase in the effective sludge activity in the mixed liquor by reducing or eliminating anaerobic periods such as can occur in clarification.

McKinney and Pfeffer (6) more recently reviewed the use of oxygen in activated sludge. Increased metabolism rates, produced by eliminating periods of zero dissolved oxygen, would increase treatment efficiencies in overloaded plants and reduce the size required for new plants. Thus, potential reductions in capital investment were viewed possible for oxygen systems.

Union Carbide recently developed the UNOX System/7 which is an oxygen aeration-activated sludge system with an oxygen utilization of over 90%. This oxygen-activated sludge process (UNOX) is presently being piloted in several locations. The Environmental Protection Agency, Washington, D.C. Pilot Plant has been evaluating the oxygen-activated sludge process since May of 1970. The objectives of the study are to determine process performance and operating requirements on the D.C. primary wastewater, an average domestic metropolitan wastewater.

SECTION IV

ANALYTICAL PROCEDURES

To evaluate the oxygen-activated sludge process, appropriate samples were manually composited over a 24-hour period. Samples were stored at $3^{o}C$ to minimize biological activity.

The 5-day biological oxygen demand (BOD) of the composite samples was determined by the probe method (9); the ammonia and nitrate-nitrite on a Technicon Automatic Analyzer (9)(10). The total organic carbon (TOC) was measured on a Beckman Carbonaceous Analyzer (11). The total phosphorus was determined by the persulfate method (12). All other analyses employed Standard Methods (13). Soluble phosphorus and soluble BOD were filtered through a standard glass suspended solids filter before analyses.

SECTION V

REACTOR

The first and most unique aspect of the system (Figure 1) is the gas tight biological reactor shown in Figure 2. In the EPA-DC Pilot Plant, primary effluent from the District of Columbia's plant is fed to the oxygen reactor either on steady state flow or on predetermined daily cycle (diurnal variation), normally with a 2.3:1 peak to minimum (45-105 gpm) daily flow variation (Figure 3).

Using all four available stages, the 8,100 gallon District of Columbia oxygen reactor provides 1.95 hours of detention time at the nominal influent flow of 100,000 gpd. At the peak daily flow, the detention time is 1.29 hours. Using three of the available four stages, the detention times are reduced to 1.50 hours and 1.00 hours, respectively, at the nominal and peak daily flows.

The reactor is sealed to prevent loss of oxygen and includes submerged hydraulic entrances and exits as well as simple water-sealed mixing equipment. Internal spray equipment using tap water is provided to suppress foam. Also, a partially submerged baffle plate before the internal exit trough retains the foam until the baffle plate is raised to allow the foam build-up to escape. The reactor is staged to provide the proper tank geometry for efficient mixing and oxygen usage.

Efficient oxygen usage is achieved by co-current contacting of the mixed liquor and oxygen gas through the various stages. The addition of pure oxygen to the reactor is controlled by a pressure regulator. An inlet oxygen control valve actuated by a pressure regulator maintains the overhead gas at a selected pressure usually between 1" and 4" of water. Even with large instantaneous fluctuations in oxygen consumption, the oxygen control valve maintains the selected pressure. The overhead gas pressure is normally selected to maintain the oxygen concentration at approximately 50% in the exhaust gas from the last reactor stage. Pure oxygen is introduced to the first stage where the peak oxygen demand occurs. As the oxygen is used in biological metabolism, respirated carbon dioxide and stripped inert gases reduce the oxygen concentration in the overhead gas flowing co-currently with the mixed liquor through the succeeding stages. The successive decrease of both oxygen availability and oxygen demand produces efficient oxygen use before the residual gas is exhausted from the reactor.

Mixed liquor dissolved oxygen levels in the District of Columbia oxygen reactor are held between 4.0 and 8.0 mg/l by adjusting the recirculation rate of the oxygen gas within the individual stages. The compressor in each stage pumps the overhead gas through the rotating

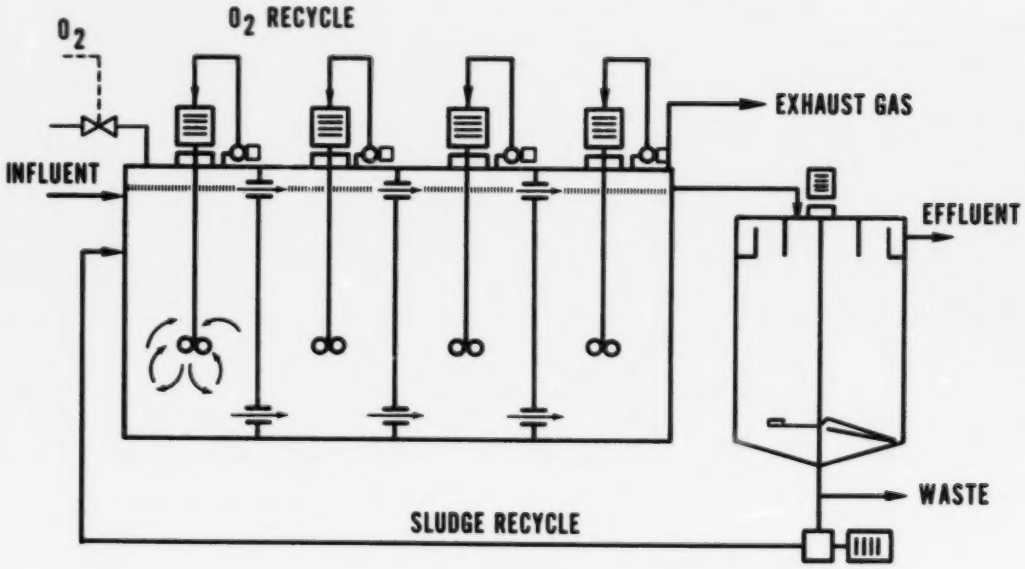

FIGURE 1
OXYGEN AERATION SYSTEM

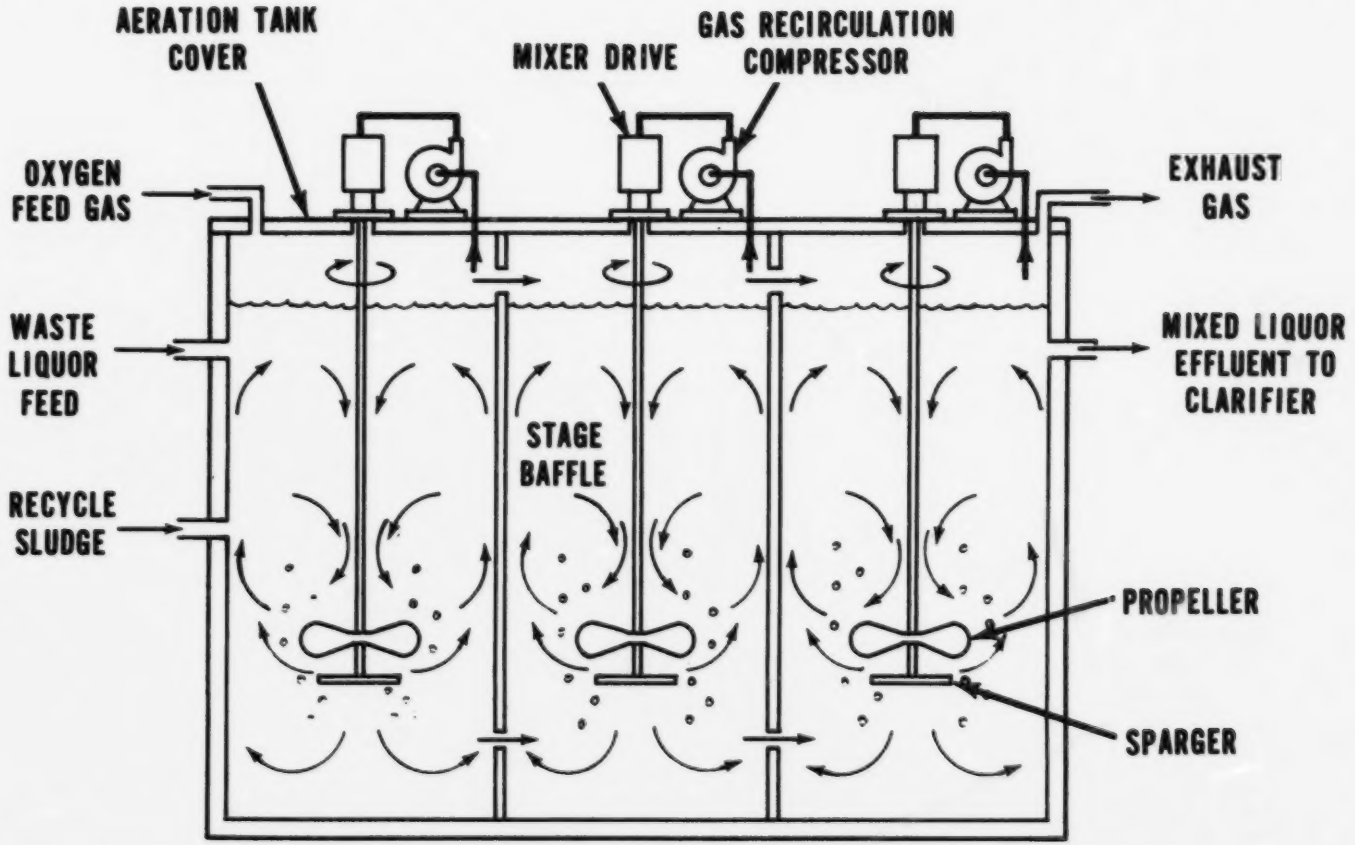

Figure 2. Oxygen Aeration Reactor

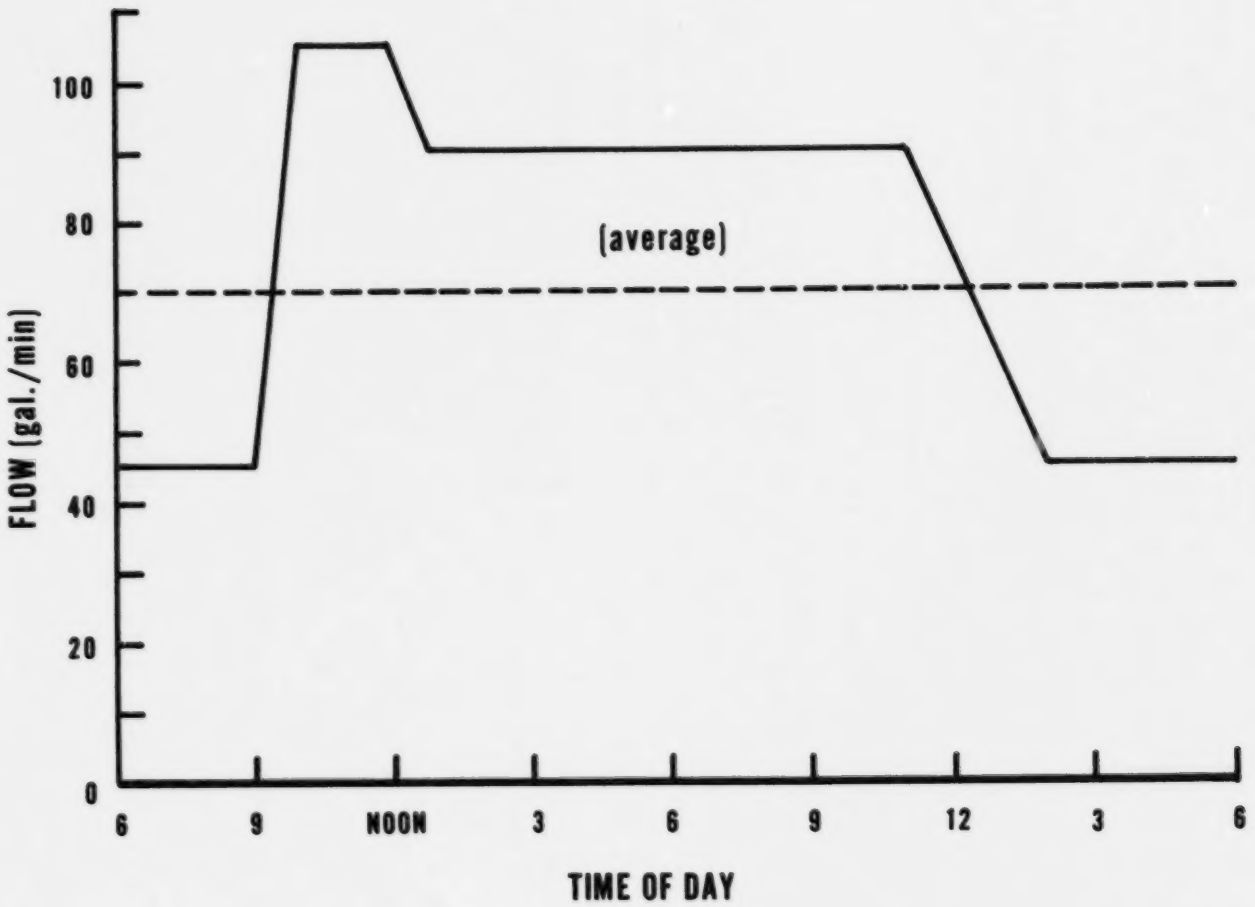

Figure 3. Diurnal Flow Variation

submerged turbine-sparger to provide efficient dispersion and mixing of the recirculated gas. The recirculation rate in each stage may be set either manually on the basis of the dissolved oxygen analysis or automatically using a control system with a dissolved oxygen sensor. The gas recirculation rate in the first District of Columbia stage typically is 3-7 cfm and 1-2 cfm in each of the last three stages. Total recirculation requirements vary between 0.10 and 0.20 ft^3/gal. of flow.

With its high oxygen transfer capabilities (which are essentially independent of turbine mixing rates), the oxygen system is able to operate at higher MLSS concentrations. These factors enable the system to readily adsorb shock organic loads. Also, toxic shock loads can be better handled, much as in a totally mixed activated sludge system. Both types of systems initially expose the toxic substrate to a large mass of active solids and the resulting "biological inertia" buffers the toxicity.

On the District of Columbia wastewater, as seen in Figure 4, the volatile portion of the oxygen solids exhibit a much higher activity for the SRT range above 6 days than the District's step aeration pilot process. The F/M ratio is the pound of BOD applied per day per pound mixed liquor volatile suspended solids (MLVSS) under aeration. Figure 4 indicates that a lower total volatile mass under aeration is required with oxygen than with air to obtain any given SRT above 6 days for a similar influent BOD. Thus, shorter detention times are required with oxygen than with step aeration at similar MLSS concentrations to achieve any given SRT above 6 days. Further, at identical SRT's above 6 days, the oxygen system will produce less excess biological solids. The most probable reason for the increased activity is attributed to maintaining the mixed liquor dissolved oxygen between 4 and 8 mg/l. The independently controlled mixing also minimized sludge pockets, dead spots and shearing of the floc particles. Mixed liquor entering the clarifier has a high dissolved oxygen content which permits a certain amount of aerobic metabolism in the clarifier and greatly reduces the time that the bio-mass is in an anaerobic condition.

The total production of solids in the oxygen system (Figure 5) per pound of BOD added, including underflow waste and effluent solids, is inversely related to the solids retention time (SRT) above an SRT of 1.3 days. The solids production with the oxygen aeration system is significantly lower than in the conventional step aeration pilot system above an SRT of 6 days. Indeed, the total solids production decreases from 0.65 pounds of excess solids per pound of BOD added at an SRT of 6 days to 0.35 pounds of excess solids per pound of BOD added at an SRT of 13 days with only a 33% increase in volatile solids concentration at the higher SRT.

The parallel conventional system, operated as step aeration or contact stabilization, exhibits increased solids production through an SRT of 9.5 days with a peak solids production of approximately 1 pound of

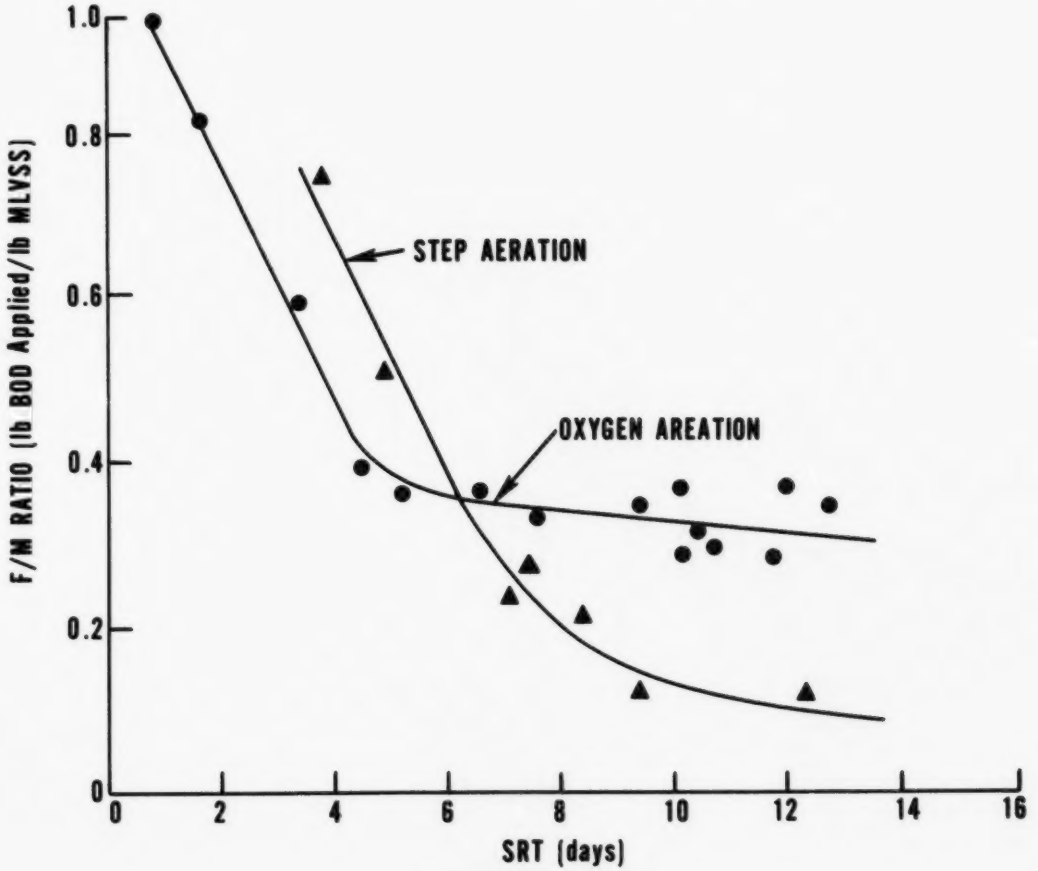

FIGURE 4
BIOLOGICAL ACTIVITY RELATIONSHIPS

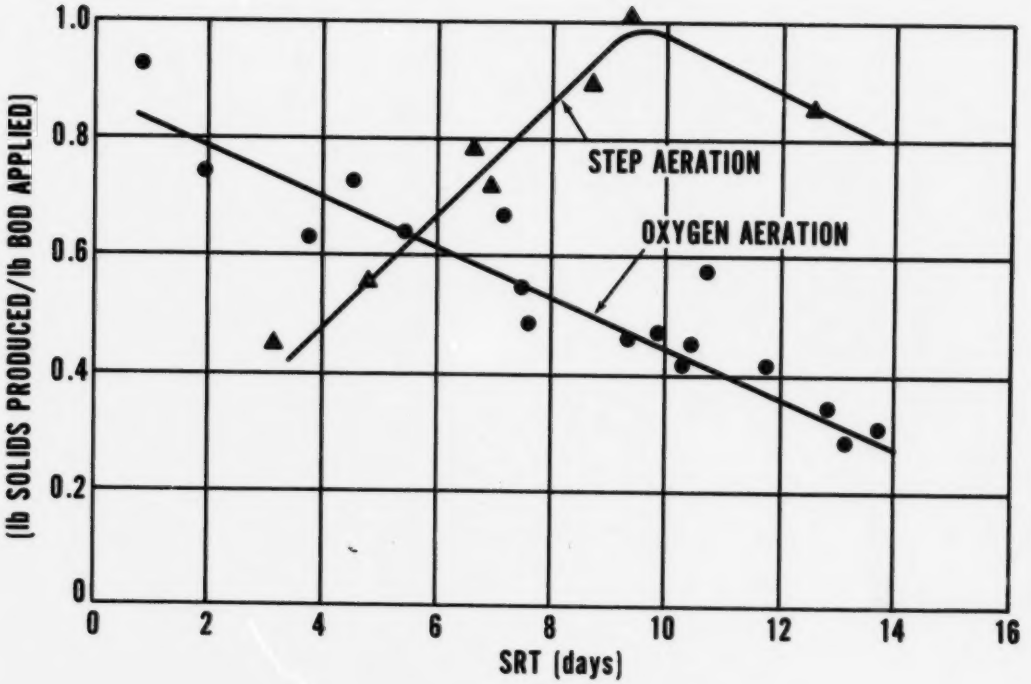

FIGURE 5
EXCESS BIOLOGICAL SLUDGE PRODUCTION

excess solids per pound of BOD added. In addition, an approximate four-fold increase in volatile solids is required to raise the SRT from 6 to 13 days in the step aeration system to achieve reduced solids production.

The reduction of BOD in the oxygen reactor is excellent. With an influent BOD up to 130 mg/l, a wide range of detention times from 1.5 to 2.5 hours, and SRT's that vary from 13 to as low as 2 days, the effluent soluble BOD is consistently less than 5 mg/l as described in Table 1. This indicates virtually complete insolubilization of the BOD in the primary effluent. Thus, BOD removal on the D.C. oxygen system is a function of clarification.

The oxygen mixed liquor is similar visually to the micro-organisms in conventional activated sludge (Figure 6). The mixed liquor biota is normally very well bioflocculated with active stalked cilates growing on the bacterial mass. Zooflagellates and free swiming cilates, although few in number, remain adjacent to or within the flocculated particles. Several varieties of large active rotifers are present in abundance. A few nematodes exist in the sludge. Normally, filamentous growth is not apparent. There is almost complete absence of fragmented debris or unflocculated bacteria between the discrete particles.

In the SRT range less than 5 days, both the oxygen activated sludge and the conventional aeration systems exhibited filamentous growth on the District of Columbia wastewater. Filamentous growth does not occur during operation above an SRT of 5 days. Normally, when encountering filamentous growth for a few days, reducing system influent flow to increase the SRT reestablishes a filamentous free sludge in several days. However, after extended periods of operation with filamentous growth, the Sphaerotilis becomes firmly entrenched and can not be quickly purged from the system by flow reduction techniques. Hydrogen peroxide added to the recycle in two 24-hour periods approximately a week apart at dosages of 200 mg/l (based on influent flow) is then required to purge the system of filamentous growth.

The recirculation of respirated carbon dioxide within the oxygen reactor stages lowers the wastewater pH from 7.0-6.8 in the first stage and to (6.1-6.4) in the final stage. With an average system pH of approximately 6.5, the oxygen process more slowly establishes a nitrifying population than the step aeration activated sludge system operated at a pH of 7.0 to 7.4. However, during the warmer months when the solids wasting is reduced to a level where the nitrifying organisms propagate faster than they are removed, the Nitrosomonas and Nitrobacter populations increase and substantial nitrification occurs in the oxygen system. Nitrogen removal across the oxygen system during periods of high nitrification and partial denitrification is as high as 39-40%. Nitrogen removal decreased to a low of 9-10% during periods without nitrification.

TABLE 1

ORGANIC REMOVAL

Operating Period Month Dates	1 June 12-30	2 July	3 August 1-25	4 September	5 October 3-11	6 November	7 January 1-16
Primary Effluent BOD (mg/l)	89	87	89	106	116	131	124
Final Effluent BOD (mg/l)	18	19	12	13	14	27	11
Final Effluent Soluble BOD (mg/l)	--	--	2	2	3	3	3
Primary Effluent COD (mg/l)	250	244	245	252	284	275	250
Final Effluent COD (mg/l)	45	70	49	51	51	63	59
Primary Effluent TOC (mg/l)	75	65	77	100	106	91	83
Final Effluent TOC (mg/l)	14	24	15	15	15	21	21
Primary Effluent Suspended Solids (mg/l)	113	101	102	107	120	92	98
Final Effluent Suspended Solids (mg/l)	36	53	28	24	35	56	24

TABLE 1 (CONTINUED)

ORGANIC REMOVAL

8 January 17-31	9 March 1-18	10 April	11 May	12 June	13 July	14 August	15 September	16 October	17 November	18 December 1-21	19 December 22-31
134	121	107	140	110	129	110	149	120	125	125	135
32	27	10	7	8	14	15	15	14	20	20	18
3	--	4	4	--	--	--	4	4	5	5	5
256	251	267	278	238	235	219	239	224	244	238	236
99	76	48	51	45	35	32	35	37	54	59	53
87	88	81	92	74	78	69	79	69	75	76	91
26	22	17	18	18	14	13	15	14	17	19	19
100	104	83	120	100	103	97	95	81	90	95	95
58	49	18	12	13	11	16	15	15	23	23	18

FIGURE 6 PHOTOMICROGRAPH OF THE MIXED LIQUOR

As in the parallel step aeration process, nitrification in the oxygen system begins to decrease in the fall and becomes virtually nil during the winter. At wastewater temperatures of about 63°F, 5 mg/l of NO_3-N is still produced with operation of an SRT of 9.0 days.

Without alum addition, phosphorus is removed from the oxygen system through metabolic uptake and by wasting of excess solids; thus, the removals vary with the metabolism of the mixed liquor. At high SRT's (highly endogenous metabolism), total phosphorus removal averages only about 15%. At lower SRT's with less endogenous respiration, phosphorus removals increase to 20%.

With alum addition, phosphorus removal in the oxygen system increases as the alum weight ratio (Al^{+++}/P) increases. During experiments conducted in the fall of 1970, for a dosage equal to an Al^{+++}/P weight ratio of 1.4/1, 80% of the phosphorus was removed to an average residual of 1.8 mg/l as P and only a slight decrease in wastewater alkalinity and pH occurred. The filtered effluent (though 0.45 u) contained an average of 1.6 mg/l of soluble P. When the dosage was increased to a ratio of 1.85/1 (Al^{+++}/P), the residual total and soluble phosphorus decreased to 0.62 and 0.53 mg/l as P, respectively. At this higher dosage, however, the buffering capacity of the oxygen mixed liquor was further reduced and the average pH decreased from 6.5 to 6.0. The oxygen biomass dispersed, necessitating termination of the alum addition to allow the mixed liquor to recover. In areas with low alkalinity wastewaters, additional alkalinity in the form of lime or caustic may be required to control pH at a level which will prevent floc dispersion. This pH adjustment may be necessary in either air or oxygen systems but is more likely in an oxygen system because of the increased dissolved CO_2 content of the mixed liquor. The addition of alum and precipitation of $Al(PO_4)$ and $Al(OH)_3$ increases the inert solids carried in the system and adequate clarification capacity for the higher solids concentration must be provided.

Consistently throughout the operation, vented gas from the fourth oxygen reactor stage has been less than 10% of the input oxygen volume. The vented stream is roughly 50% oxygen. Based upon the influent and exhausted oxygen concentrations, the net utilization of oxygen in the process is about 95%. The accountable oxygen consumption consisting of COD removal, nitrification demand, exhaust gas, and effluent dissolved oxygen is summarized in Table 2. The COD removed was calculated by subtracting the COD in the underflow waste solids and that in the process effluent from the primary effluent COD. With increasing SRT, additional oxygen is required for endogenous respiration. Likewise during periods with nitrification, additional oxygen is required.

Pertinent reactor variables and operating conditions are summarized in Table 3 for approximately 1-1/2 years of operation on the District of Columbia oxygen aeration pilot plant.

TABLE 2 - OXYGEN USAGE

Operating Period Month Dates	1 June	2 July	3 August	4 September	5 October 3-21	6 November 10-30	7 January 1-16
Primary Effluent COD (lb/million gal.)	2080	2030	2040	2090	2370	2290	2080
Final Effluent COD (lb/million gal.)	375	584	408	430	425	525	488
Waste Sludge COD (lb/million gal.)	188+	42+	150+	217+	630+	247	233
COD Removed from System (lb/million gal.)	1517	1404	1482	1443	1315	1518	1359
Nitrate Nitrogen Demand (lb/million gal.)	14	69	128	160	100	18	9
Exhaust Oxygen (lb/million gal.)	85	54	75	85*	87	85	65
Final Effluent D.O. (lb/million gal.)	10*	10*	10	10	10	10*	25
Total	1626	1537	1695	1698	1512	1631	1458
Oxygen Supplied (lb/million gal.)	1750	1825	1775	1900	1650	±	1700**

\+ COD = 1.4 volatile solids

* Estimate

± Inlet meter malfunctioned

** Increase sampling and greater losses of O_2 through sample ports

TABLE 2 - OXYGEN USAGE - CONT'D

8 January 17-31	9 March 1-18	10 April	11 May	12 June	13 July	14 August	15 September	16 October	17 November	18 December 1-21	19 December 22-31
2140	2030	2180	2500	1750	1940	1830	1920	1880	1950	1990	1970
826	600	380	450	340	270	260	280	310	440	490	440
336	160	700	890	80	270	270	400	390	220	230	300
978	1260	1100	1100	1330	1400	1300	1240	1180	1290	1270	1230
14	0	0	0	0	60	70	270	200	220	190	70
65	160	130	40	80	300	260	200*	200	200	200	150*
25	30	40	50	60	60	40	40	30	40	40	40
1082	1450	1270	1250	1470	1820	1670	1750	1610	1750	1700	1490
1800**	1450	1300	1260	1600	2200	1690	±	±	1740	2000	1450

TABLE 3

REACTOR VARIABLES AND OPERATING CONDITIONS

Operating Period Month Dates	1 June 12-30	2 July	3 August 1-25	4 September	5 October 3-21	6 November 10-30	7 January 1-16
Flow Rate (gpm)	50-55	80	80	70+	70+*	70+	53
Aeration Time (hr.)	2.00	1.66	1.66	1.95	1.95	1.95	2.50
Recycle Rate	50	50	42	32	38	37	77
MLSS (mg/l)	4140	5180	5250	6000	8120	6350	5300
MLVSS (%)	74	70	73	78	67	73	80
SRT (days)	7.7	7.3	11.8	10.7	5.5	5.5	13.0
F/M (lb BOD/day/lb MLVSS)	0.333	0.342	0.296	0.304	0.283	0.355	0.275
Volumetric Loading (lb BOD/ day/1,000 ft^3)	57	80	80	96	106	108	89
Mixer Power $(\frac{k.w. - hr.}{1,000\ gal.})$**	1.27	1.19	0.98	0.92	1.00	1.00	1.18
Conpressor Power $(\frac{k.w. - hr.}{1,000\ gal.})$**	0.39	0.28	0.39	0.41	0.35	0.26	0.34
Temperature (^oF)	74-80	78-84	82-85	79-83	70-79	56-69	58-60

* Alum addition
\- 2.3:1 diurnal variation
** Pilot plant equipment efficiency was not determined

TABLE 3 (CONTINUED)

REACTOR VARIABLES AND OPERATING CONDITIONS

8 January 17-31	9 March 1-18	10 April	11 May	12 June	13 July	14 August	15 September	16 October	17 November
53	60-70	31-67	60	30-70	70	70+	70+	70+	70+
2.50	2.15	3.30-1.55	1.70	3.70-1.50	1.5	1.5	1.5	1.5	2.0
80	50-60	90-60	65	100-50	50	50	46	46	45
3940	8070	2710	2750	1000	6600	7500	7400	6000	4600
81	77	81	78	73	70	70	72	73	78
4.7	3.7	1.3-4.0	2.0	13.0	12.6	10.0	7.5	9.5	9.8
0.392	0.580	0.30-1.00	0.970	0.400	0.430	0.32	0.39	0.31	0.39
80	90	98	157	95	160	131	185	146	111
1.42	-	-	-	-	-	-	-	-	-
0.32	-	-	-	-	-	-	-	-	-
58-60	50-62	62-65	65-71	70-77	77-80	77-81	76-81	76-71	71-65

TABLE 3 (CONTINUED)

REACTOR VARIABLES AND OPERATING CONDITIONS

18 December 1-21	19 December 22-31
70+	70+
2.0	2.0
40	40
4400	4200
80	81
9.0	6.5
0.40	0.50
111	122
-	-
-	-
65-63	63-61

SECTION VI

CLARIFICATION

The second important aspect in the oxygen aeration system is the method of liquid/solids separation. As mentioned before, soluble residual BOD in the effluent averaged less than 5 mg/l in the test periods indicating virtually complete BOD insolubilization. Thus, most of the residual BOD in the District of Columbia oxygen system effluent is associated with suspended solids. Overall removal of suspended solids and its associated BOD is a functional of clarification efficiency.

Clarifier efficiency is in turn a function of the basic settling characteristics of the solids as well as of the actual design and operation of the clarifier. With the normally higher mixed liquor concentrations used in the oxygen aeration process, design criteria for both clarifiers (i.e., overflow rates and volume) and thickeners (i.e., solids loading - $lb/ft^2/day$) should be considered. The Ten State Standards suggest that conventional activated sludge clarifiers be designed for average overflow rates of 800 gpd/ft^2. The Water Pollution Control Federation Manual of Practice (1959) suggests that the solids loading be held below a peak of 30 $lb/ft^2/day$. Overflow rates and solids loading criteria should be better defined for high solids systems such as oxygen aeration.

The basic settling characteristics of the mixed liquor typically have been found to be a function of:

1. Concentration of the mixed liquor.
2. Particle shape.
3. Particle density.
4. Seasonal variation
    a. Physical changes in water density and viscosity with temperature.
    b. Metabolic changes with temperature.
    c. Seasonal loading variation.

As seen in Figure 7, the log of the initial settling rate is a function of the log of the solids concentration. This curve illustrates that two relationships exist. The first at lower MLSS levels corresponds to free particle settling and is characterized by the absence of an initial discrete subsiding interface and a zone of homogenous settling solids. The second at higher MLSS levels had both an initial descrete interface and a zone of homogenous settling particles (zone settling) (14). Thus, the sizing of a clarifier is a function of the MLSS concentration and must be coordinated with the reactor (and the sludge

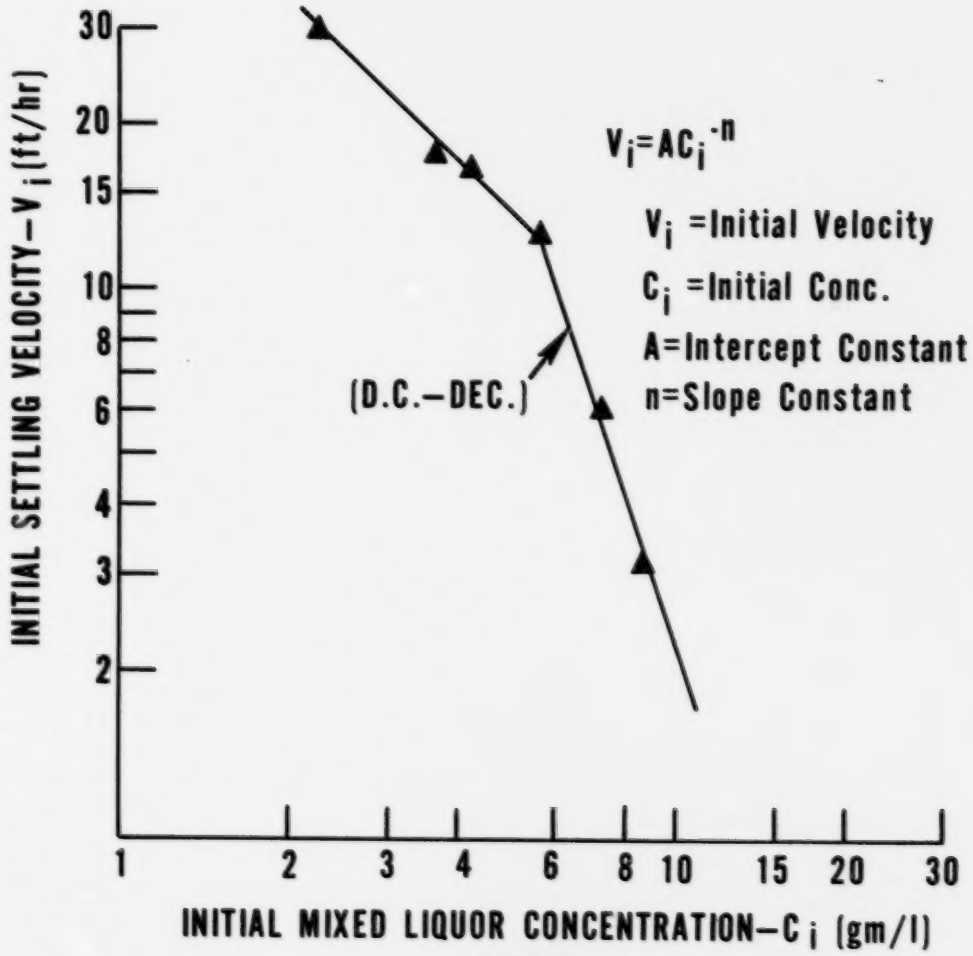

FIGURE 7

INITIAL SLUDGE SETTLING VELOCITY PROFILE FOR BROAD MIXED LIQUOR CONCENTRATION RANGE

handling facilities) to achieve the desired biological capabilities of the system.

Another important factor is the particle shape. Normally, as shown in Figure 6, the oxygen mixed liquor particles have rounded shapes. However, if filamentous growth exists, as experienced below an SRT of 5 days in D.C. (air and oxygen), both settling rates and compaction deteriorate. The range, if any, that filamentous growth appears is unique to each location and should be defined for that location.

Still another important factor in the basic settling characteristics is the density of the particles in relationship to the density of the water. It is the difference in density that is the driving force for settling. The VSS/TSS ratio (or volatile %) is one relative indication of density. There are several ways to improve the particle density. One is to feed raw wastewater instead of primary effluent to the oxygen aeration system thus incorporating the normally denser particles captured in primary sedimentation into the biomass, such as occurred at Batavia (7). Again, the sizing of the reactor oxygen supply, etc., must be compatible with the increased organic loading. In Washington, heavy rains and unusually high flows wash silt and clay into the sewer system. These materials subsequently become incorporated in the mixed liquor solids and have increased sludge settling rates 30% to 60%. Also, operation under different biological conditions can alter sludge settling characteristics.

Another unique method of increasing the density of the sludge that was evaluated at the EPA-DC Pilot Plant is by altering the method of clarifier operation. Two major methods of clarifier operation are possible. One is to use the blanket as a filter and the other is to permit classification of the settling solids. The first method can be accomplished in two different ways: (1) by providing sufficient depth to the clarifier such that the mixed liquor passes up through the clarifier blanket (which acts as a filter) (2) by carrying high MLSS concentrations (usually above 4500 mg/l in D.C.) such that the particles settle in a subsidence (zone) settling pattern with discrete interfaces existing between the homogeneous subsiding particles and the decant. In the subsidence zone, the relatively uniform concentration of particles are nearly homogeneously mixed by the countercurrent turbulence produced by water passing around the solids. The homogeneous subsiding blanket does not allow classification of individual particles because the settling blanket acts as a filter.

At the District of Columbia, with MLSS concentrations below 4500 mg/l, subsidence (zone) settling does not occur during the initial portion of settling. This provides for a second method of clarifier operation where classification of the discrete settling particles can occur if the mixed liquor is fed above the clarifier blanket level. The lighter or unsettable particles, thus, can be purged from the system.

The effluent suspended solids accordingly increased from 15 mg/l to 25 mg/l during this method of operation with a corresponding increase in effluent BOD.

Seasonal variations also affect sludge settling characteristics in oxygen as well as in air systems. These variations become critical as the concentration of solids increases. The pure physical changes in the wastewater density and viscosity contribute to slower settling rates as the wastewater temperature decreases. As the density of the water increases, the driving force for settling (which is the difference in density between water and the settling particles) decreases for a similar particle density. The drag force, viscosity, also increases with decreasing temperature ($\sim$25% from 80^{o}F to 55^{o}F) again contributing to slower settling rates in colder waters. Figure 8 shows a series of liter batch settling tests conducted in June, 1971 by only altering the temperature of the mixed liquor (unacclimated biota). As expected, the colder samples settled slower. These tests are compared in Figure 8 to a similar series of tests in which the mixed liquor was acclimated to the colder wastewater temperature of January, 1971. In Figure 9, the batch flux (concentration multiplied by settling velocity) or the solids loading in lb/ft^{2}/day is shown for the previous tests. Again, the effect of wastewater temperature is evident.

Besides the physical changes caused by seasonal variations, another factor which must be considered is the metabolic change brought about by changing wastewater temperature. Figure 10 shows that the settling characteristics of oxygen mixed liquor change seasonably at D.C. The clarifier was operated as a sludge blanket to capture the normally unsettleable solids. At similar SRT's, the initial settling rate in a 1 liter graduated cylinder test decreased from approximately 10 ft/hr to 7 ft/hr at a concentration of 6000 mg/l as the temperature changed from 81^{o}F to 71^{o}F. In Figure 11, with the clarifier operated as a slurry pool to purge the normally unsettleable particles; the solids also showed a decreasing initial settling rate with decreasing wastewater temperature for a similar biology. The initial settling rate in the 1 liter test decreased from 14 ft/hr to 9 ft/hr at 4500 mg/l as the temperature decreased from 70^{o}F to 63^{o}F. Similar patterns of decreasing settling rates with decreasing wastewater temperature have been observed in nitrifying and denitrifying mixed liquors also.

Clarifier operation and design are equally important to the basic settling characteristics of the solids in gravity clarification. Besides selecting an overflow rate compatible with reactor sizing, the depth and method of clarifier feed are important as discussed earlier for either high solids capture or solids classification. Another important design consideration is the volume or detention time of the clarifier. At the District of Columbia, the oxygen system underflow

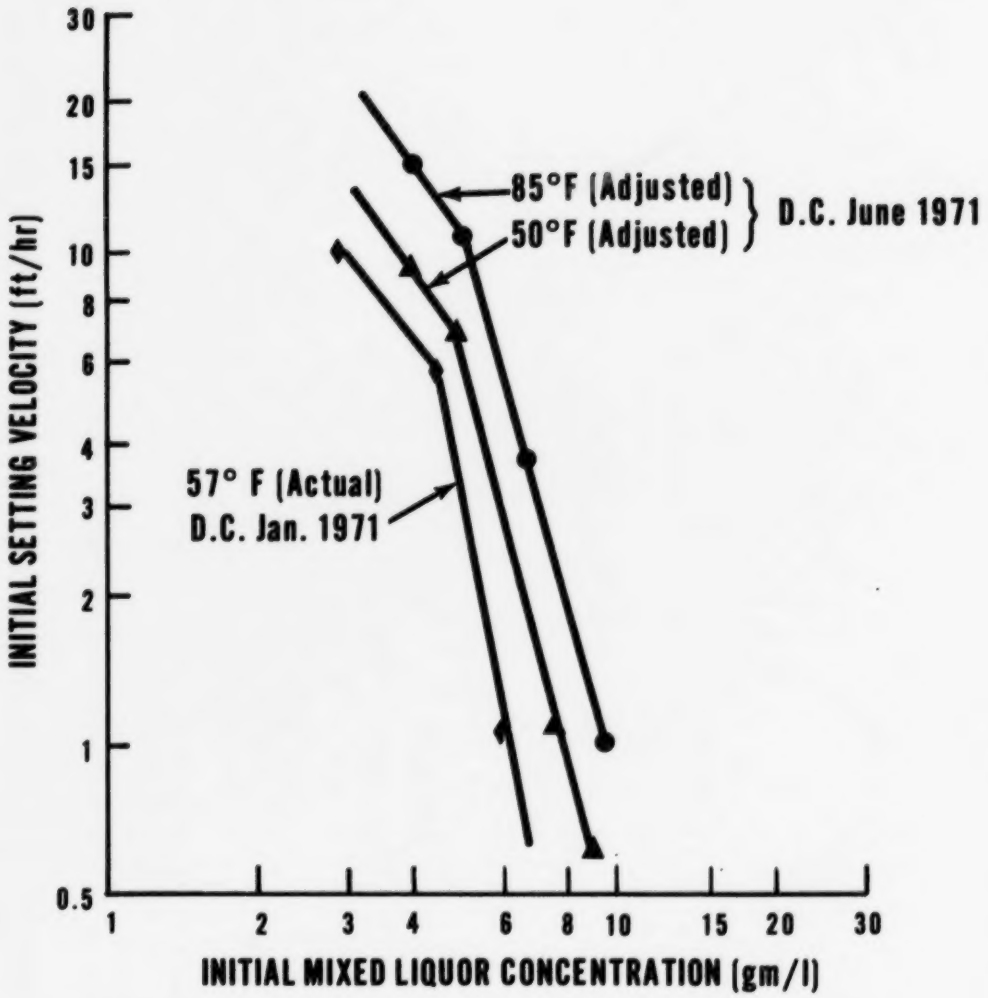

Figure 8. Effect of Adjusted vs. Acclimated Wastewater Temperatures on Sludge Settling Rates

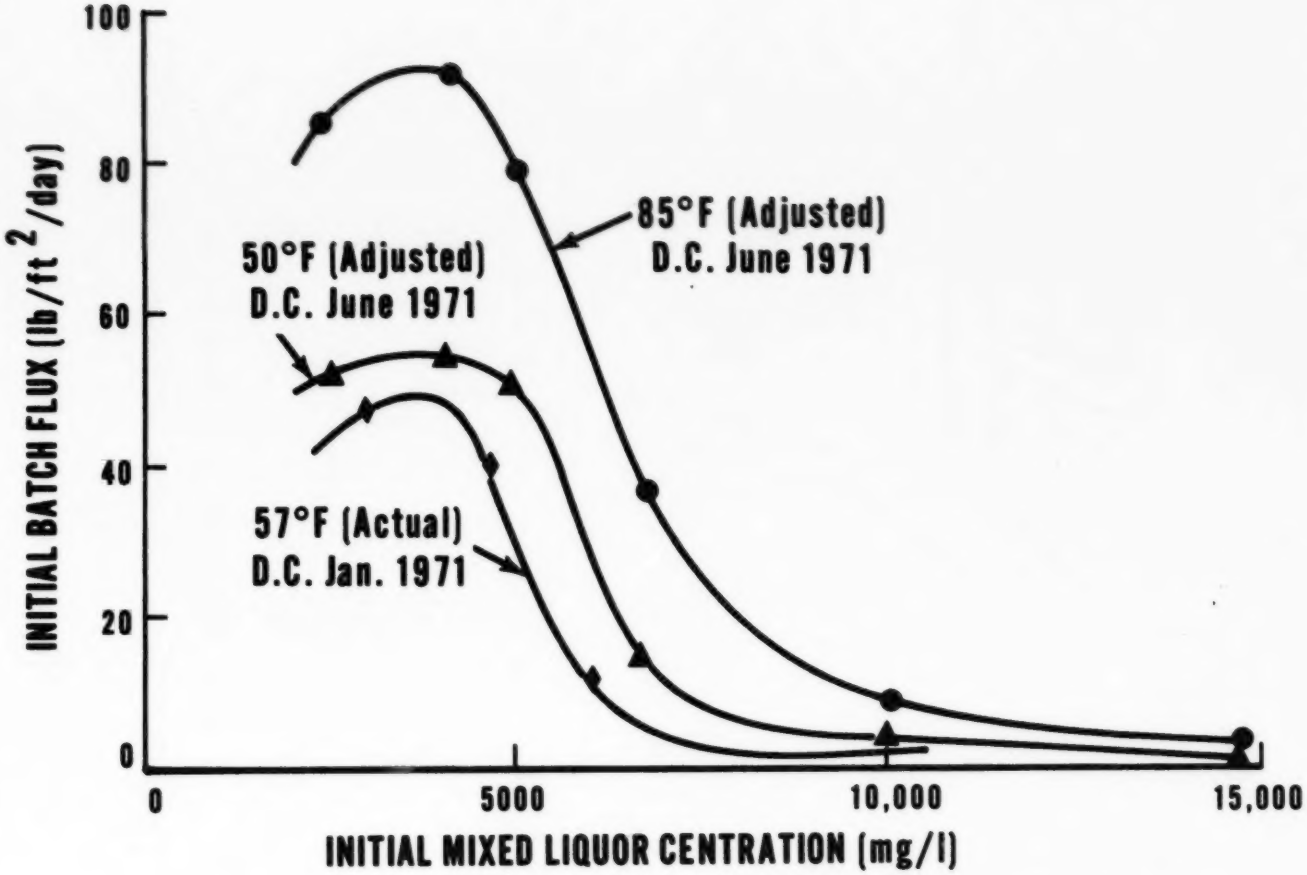

Figure 9. Effect of Wastewater Temperatures on Initial Batch Flux

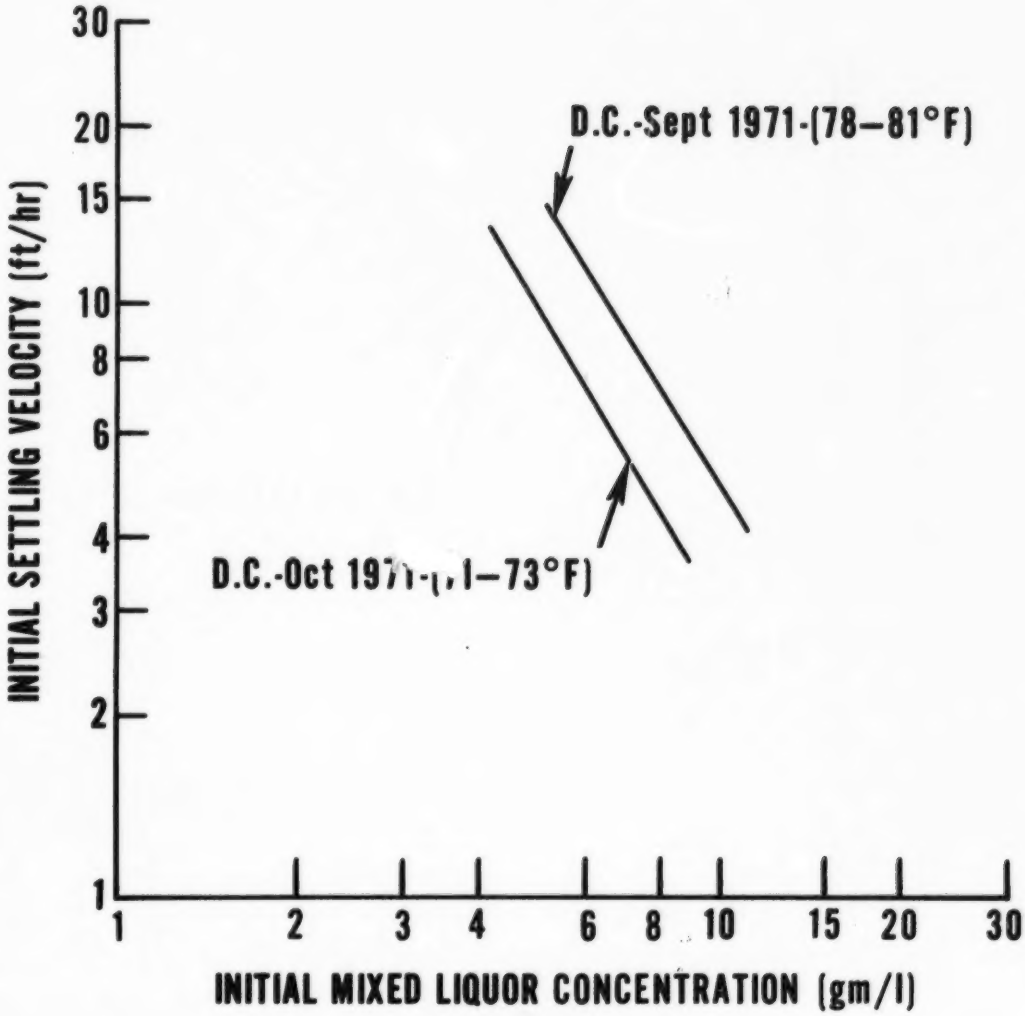

FIGURE 10

EFFECT OF WASTEWATER TEMPERATURE ON OXYGENATED SLUDGE SETTLING RATES

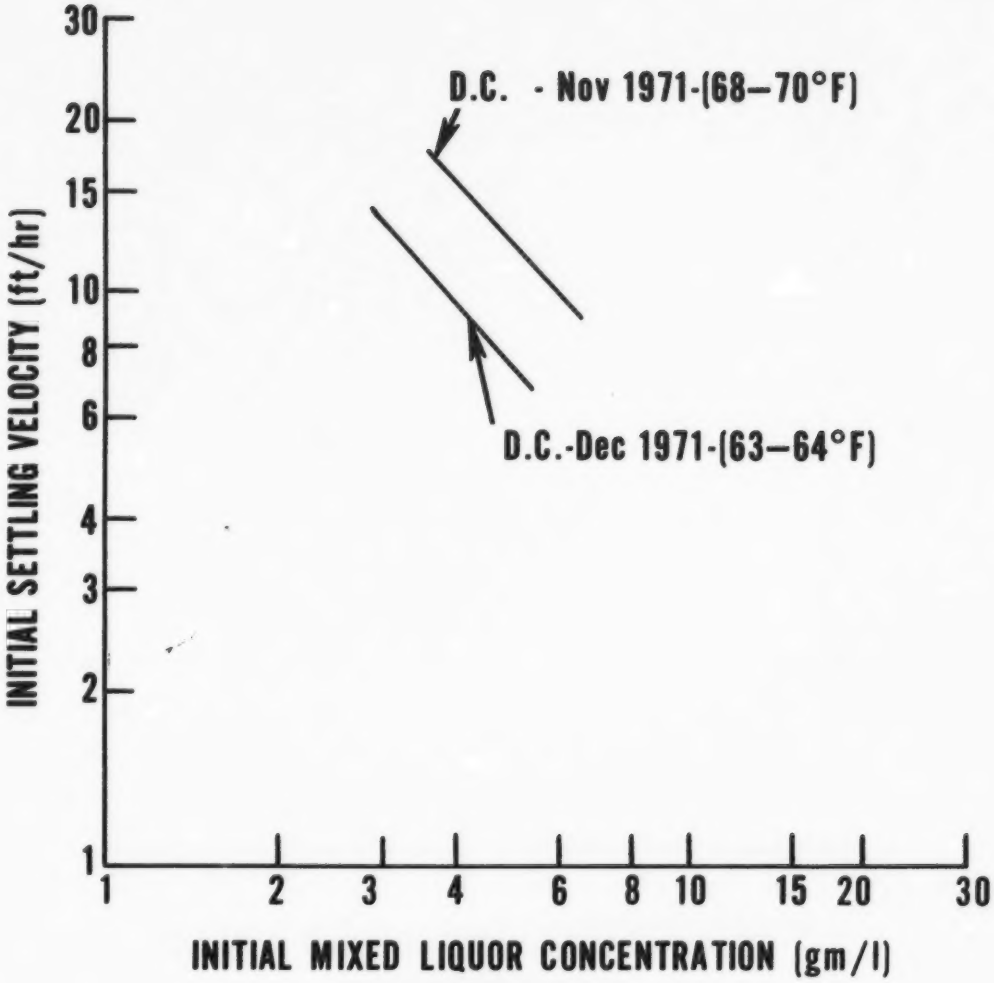

FIGURE 11

EFFECT OF WASTEWATER TEMPERATURE ON OXYGENATED SLUDGE SETTLING RATES

solids concentration varies between 1.0% and 1.4% with an average clarifier detention time of 1.9 hours. With 2.8 hours average detention time, the underflow solids concentration rises to 2.0%-2.4%. The sludge recycle rate is then determined after an F/M ratio is established for the reactor and the underflow concentration from the clarifier is likewise established.

Inventory solids are another consideration in clarifier operation. The total solids inventory is a result of both the build-up of solids and the solids actually in the transition (or settling) process. The build-up of solids can be removed by simply increasing the underflow wasting. However, as the settling rates of the mixed liquor decrease for a given concentration (i.e., with temperature) an increase in inventory results as more solids are required in transition (or settling).

In the summer of 1970 at an MLSS concentration in excess of 8000 mg/l, peak clarifier overflow rates of 1940 gpd/ft^2 were observed on the District of Columbia oxygen system as shown in Table 4. During the 1970-71 winter, the peak sustained overflow rates which could be maintained without the blanket coming over the weirs was 975 gpd/ft^2 at MLSS concentrations that varied from 3900 to 5300 mg/l. The causative agents which reduced the allowable overflow rates for satisfactory operation from summer to winter were undoubtedly a combination of all the above mentioned factors, not the least of which was the decreased wastewater temperature. As expected with filamentous growth in late spring 1971, allowable clarifier overflow rates were markedly decreased as shown in Table 4. Again, during the 1971-72 winter, maximum overflow rates of 975 gpd/ft^2 have been demonstrated at the District of Columbia.

TABLE 4

CLARIFIER VARIABLES AND OPERATING CONDITIONS

Operating Period	1	2	3	4	5	6	7	8
Month	June	July	August	Sept.	Oct.	Nov.	Dec.	Jan.
Dates	12-30		1-25		3-21	10-30	1-16	17-31
Average Overflow Rate (gpd/ft^2)								
At Surface	- +	-	- +	1280	1280±	1280±	975±	975±
Above Feed Skirt	750	1210	1210	1050	1050	1050	800	800
Below Feed Skirt	670	1075	1075	940	940	940	710	710
Peak Overflow Rate (gpd/ft^2)								
At Surface	-	-	- +	1940	1940	1940	975	975
Above Feed Skirt	750	1210	1210	1580	1580	1580	800	800
Below Feed Skirt	670	1075	1075	1410	1410	1410	710	710
Average Solids Loading ($lb/day/ft^2$)	37	75	58	61	88	68	55	42
SVI	80	48	50	42	33	48	60	73
Underflow Solids								
% Dry Solids	1.16	1.34	1.27	1.40	2.14	1.40	1.08	1.00
% Volatile	75	70	75	80	65	81	90	80
Waste Solids (lb/million gal.)	161	40	144	250	680	230	193	253
Volatile (lb/million gal.)	121	28	108	200	441	202	174	202
Effluent Solids								
(lb/million gal.)	296	445	166	204	290	470	197	483
Volatile (lb/million gal.)	198	245	113	141	189	342	118	400

\+ Peripheral feed - no center feed section
Area at surface 96 ft^2 x 6 ft^2 x 5 ft. deep
Area below feed skirt 107 ft^2 x 5 ft. deep
Total depth 11 ft. deep

± Center feed section: area at surface 78 $ft.^2$ x 4 ft. deep; area above feed skirt 96 ft^2 x 2 ft. deep; area below feed skirt 107 ft^2 x 5 ft. deep.

TABLE 4 (CONTINUED)

CLARIFIER VARIABLES AND OPERATING CONDITIONS

9 March 1-18	10 April	11 May	12 June	13 July	14 Aug.	15 Sept.	16 Oct.	17 Nov.	18 Dec. 1-21	19 Jan. 22-31
950±	-@	-	-	-@	-@	-@	-@	-@	-@	-@
780	(290-620)	560	(280-650)	975	975	975	975	975	975	975
700	-	-	-	-	-	-	-	-	-	-
950	-	-	-	-	-	-	-	-	-	-
780	(290-620)	560	(280-650)	650	650	650	650	650	650	650
700	-	-	-	-	-	-	-	-	-	-
35	17	21	24	54	61	58	47	36	33	32
81	120-190	265	173	50	30-35	33	42	36	40	56
0.79	0.85	0.78	1.28	1.92	2.22	2.58	2.26	2.05	2.41	2.30
77	80	77	78	70	70	70	71	79	81	82
130	-	720	142	168	253	460	340	178	200	310
100	-	550	104	118	178	323	242	140	160	254
410	-	100	88	92	133	160	130	194	190	140
375	-	-	-	-	100	104	99	130	123	88

@ Two center feed clarifiers at 78 ft^2 each x 11 feet deep
feed sections at 3 ft^2 each x 3.5 ft deep

SECTION VII

SOLIDS HANDLING

The other integral part of the oxygen system is the excess solids handling equipment. Of utmost importance is the relative ease with which the oxygen activated sludge process can be operated in endogenous respiration, thereby substantially reducing the quantity of excess sludge to be handled. This factor will reduce the number and/or size of the selected sludge handling and disposal facilities. However, the increased operating costs resulting from the increased oxygen necessary to oxidize ("burn-up") the excess sludge and the larger reactor/clarifier capabilities needed to hold the increased solids inventory required for endogenous respiration must be balanced economically with the reduction in size of the solids handling and disposal units.

Another factor to be considered is that the larger the clarifier volume, the thicker the underflow solids concentration. It may be economically feasible to properly size the reactor/clarifier combination to yield underflow solids sufficiently thick to be dewatered directly without prior thickening or digestion. This would be accomplished by selecting a small reactor and large clarifier. An alternative to the above is to select a large reactor and a small clarifier and provide additional thickener capabilities. In the EPA-DC Pilot Plant, the excess solids are thickened separately by air flotation or gravity thickening. These solids have been thickened to over 4.5% without chemical additives by both gravity and air flotation thickening.

SECTION VIII

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SECTION IX

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